CFD SIMULATION OF SWIRLING IN FLUIDIZED BED BY USING ANNULAR TYPE DISTRIBUTOR

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ABSTRACT

This paper report about the swirling fluidized bed (SFB) which is affected by the designs of perforated plate. The result of the flow simulation for the each distributor plate perforated, inclines and annular are produces by using the Solid Work Flow Simulation intuitive (CFD). The characteristic of the each design plate are different in their number of hole, diameter of hole, thickness of plate and diameter of plate in order to get the best result which respect to pressure drop. The performance of the SFB was assessed in term of pressure drop values, minimum fluidization velocity, Umf. Also the performance of the each plate are looked at their flow air pattern in fluidized bed, which are the more swirl pattern of air the more better in result. More importantly is the reduction pressure drop in the appropriate design in distributor plate. The good results in this study were produced by the annular plate which is able to produce a minimum pressure drop compared with the perforated and Incline plate. While the annular plate also shown the swirl of air pattern better than perforated and incline plate. Furthermore, to ensure better results in this study, the experiment shall be conducted so that the results of the experiment can be compared with the flow simulation results. Besides that, from the experiment also the results that produce have more actual compare with flow simulation result.

ABSTRAK

Laporan ini adalah mengenai pendiang bendaliran berpusar (SFB) yang dipengaruhi oleh reka bentuk piring berlubang. Hasil simulasi aliran bagi setiap jenis piring berlubang, cenderung dan anulus dapat dihasilkan dengan menggunakan Solid Work Simulasi Aliran intuitif (CFD). Ciri-ciri yang setiap reka bentuk piring adalah yang berbeza terhadap bilangan numbor piring lubang, diameter setiap lubang, ketebalan piring dan diameter piring. untuk mendapatkan hasil yang bagus terhadap kejatuhan tekanan. Keberkesanan SFB telah dinilai dari segi nilai-nilai kejatuhan tekanan, minimum halaju pembendaliran , Umf. Juga keberkesanan setiap piring dapat dilihat juga pada pusaran corak aliran udara di dalam pendiang bendaliran, yang mana corak pusaran lebih kuat dapat menghasilkan hasil yang lebih baik. Dalam masa yang sama perkara yang paling penting adalah penurunan tekanan dapat dihasilkan pada tahap yang paling minimum oleh setiap reka bentuk piring berlubang. Hasil yang terbaik dalam kajian ini dapat ditunjukkan oleh piring anulus yang mampu menghasilkan penurunan tekanan minimum berbanding dengan plat berlubang dan cenderung. Disamping itu, piring anulus juga menunjukkan corak pusaran udara yang lebih baik daripada piring berlubang dan cenderung. selanjutnya, bagi memastikan hasil yang lebih baik dalam kajian ini, eksperimen hendaklah dijalankan supaya hasil eksperimen boleh dibandingkan dengan hasil yang dihasilkan oleh simulasi aliran. Selain itu, dari eksperimen juga dapat menghasilkan mempunyai gambaran yang sebenar berbanding dengan hasil simulasi aliran.

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C_d	Coefficient of discharge
d	Diameter (m)
H	Angular momentum (kg m^2s^{-1}
u, U	Velocity (ms^{-1})
μ	Friction coefficient, dynamic viscosity of gas (Nsm^{-2})
ρ	Density (Kg m^{-3})
U_{mf}	Velocity minimum fluidization
U_{ms}	Velocity minimum swirl
Θ	Tangential
d_p	Particle diameter, m
Θ_s	Granular temperature of the solid, m^2/s^2
d_p	Sand particle size, μm
Δp	Pressure drop across the bed, KPa
ρ_f	Density of fluidizing (air), Kg/m^3
ρ_s	Density of the solid bed (sand) particle, Kg/m^3

CHAPTER 1

INTRODUCTION

1.1 BACKGROUND OF STUDY

In 1922 Fritz Winkler made the first industrial application of fluidization in a reactor for a coal gasification process [1]. In 1942, the first circulating fluid bed was built for catalytic cracking of mineral oils, with fluidization technology applied to metallurgical processing (roasting arsenopyrite) in the late 1940s [2][3]. During this time theoretical and experimental research improved the design of the fluidized bed. In the 1960s VAW-Lippewerk in Lunen, Germany implemented the first industrial bed for the combustion of coal and later for the calcination of aluminium hydroxide.

A fluidized bed is formed when a quantity of a solid particulate substance (usually present in a holding vessel) is placed under appropriate conditions to cause the solid/fluid mixture to behave as a fluid. This is usually achieved by the introduction of pressurized fluid through the particulate medium. This results in the medium then having many properties and characteristics of normal fluids; such as the ability to free-flow under gravity, or to be pumped using fluid type technologies.

The resulting phenomenon is called fluidization. Fluidized beds are used for several purposes, such as fluidized bed reactors (types of chemical reactors), fluid catalytic cracking, fluidized bed combustion, heat or mass transfer or interface modification, such as applying a coating onto solid items. This technique is also becoming more common in Aquaculture for the production of shellfish in Integrated Multi-Trophic Aquaculture systems. [4]

A fluidized bed consists of fluid-solid mixture that exhibits fluid-like properties. As such, the upper surface of the bed is relatively horizontal, which is analogous to hydrostatic behavior. The bed can be considered to be an inhomogeneous mixture of fluid and solid that can be represented by a single bulk density.

Furthermore, an object with a higher density than the bed will sink, whereas an object with a lower density than the bed will float, thus the bed can be considered to exhibit the fluid behavior expected of Archimedes' principle. As the "density", (actually the solid volume fraction of the suspension), of the bed can be altered by changing the fluid fraction, objects with different densities comparative to the bed can, by altering either the fluid or solid fraction, be caused to sink or float.

In fluidized beds, the contact of the solid particles with the fluidization medium (a gas or a liquid) is greatly enhanced when compared to packed beds. This behavior in fluidized combustion beds enables good thermal transport inside the system and good heat transfer between the bed and its container. Similarly to the good heat transfer, which enables thermal uniformity analogous to that of a well-mixed gas, the bed can have a significant heat-capacity whilst maintaining a homogeneous temperature field.

Fluidized beds are used as a technical process which has the ability to promote high levels of contact between gases and solids. In a fluidized bed a characteristic set of basic properties can be utilized, indispensable to modern process and chemical engineering, these properties include:

i. Extremely high surface area contact between fluid and solid per unit bed volume
ii. High relative velocities between the fluid and the dispersed solid phase.
iii. High levels of intermixing of the particulate phase.
iv. Frequent particle-particle and particle-wall collisions.

Taking an example from the food processing industry: fluidized beds are used to accelerate freezing in some IQF tunnel freezers. IQF means Individually Quick Frozen, or freezing unpackaged separate pieces. These fluidized bed tunnels are typically used on small food products like peas, shrimp or sliced vegetables, and may use cryogenic or vapor-compression refrigeration.

The fluid used in fluidized beds may also contain a fluid of catalytic type; that's why it is also used to catalyst the chemical reaction and also to improve the rate of reaction.

Bed types can be coarsely classified by their flow behavior, including [5]:

i. Stationary or bubbling bed is the classical approach where the gas at low velocities is used and fluidization of the solids is relatively stationary, with some fine particles being entrained.
ii. Circulating fluidized beds (CFB), where gases are at a higher velocity sufficient to suspend the particle bed, due to a larger kinetic energy of the fluid. As such the surface of the bed is less smooth and larger particles can be entrained from the bed than for stationary beds. Entrained particles are recirculating via an external loop back into the reactor bed. Depending on the process, the particles may be classified by a cyclone separator and separated from or returned to the bed, based upon particle cut size.
iii. Vibratory Fluidized beds are similar to stationary beds, but add a mechanical vibration to further excite the particles for increased entrainment.
iv. Transport or flash reactor (FR). At velocities higher than CFB, particles approach the velocity of the gas. Slip velocity between gas and solid is significantly reduced at the cost of less homogeneous heat distribution.
v. Annular fluidized bed (AFB). A large nozzle at the center of a bubble bed introduces gas as high velocity achieving the rapid mixing zone above the surrounding bed comparable to that found in the external loop of a CFB.

When the packed bed has a fluid passed over it, the pressure drop of the fluid is approximately proportional to the fluid's superficial velocity. In order to transition from a packed bed to a fluidized condition, the gas velocity is continually raised. For a free-standing bed there will exist a point, known as the minimum or incipient fluidization point, whereby the bed's mass is suspended directly by the flow of the fluid stream. The corresponding fluid velocity, known as the "minimum fluidization velocity" Umf. [6]

Beyond the minimum fluidization velocity ($U \geq U_{mf}$), the bed material will be suspended by the gas-stream and further increases in the velocity will have a reduced effect on the pressure, owing to sufficient percolation of the gas flow. Thus the pressure drop from for U ≥ Umf is relatively constant.

At the base of the vessel the apparent pressure drop multiplied by the cross-section area of the bed can be equated to the force of the weight of the solid particles (less the buoyancy of the solid in the fluid).

$$\Delta p_w = H_w(1-\epsilon_w)(\rho_s - \rho_f)g$$

In 1973, Professor D. Geldart proposed the grouping of powders in to four so-called "Geldart Groups". [7] The groups are defined by their locations on a diagram of solid-fluid density difference and particle size. Design methods for fluidized beds can be tailored based upon the particle's Geldart grouping: [6]

Group A For this group the particle size is between 20 and 100 µm, and the particle density is typically less than 1.4g/cm^3. Prior to the initiation of a bubbling bed phase, beds from these particles will expand by a factor of 2 to 3 at incipient fluidization, due to a decreased bulk density. Most powder-catalyzed beds utilize this group.

Group B The particle size lies between 40 and 500 µm and the particle density between 1.4-4 g/cm^3. Bubbling typically forms directly at incipient fluidization.

Group C This group contains extremely fine and consequently the most cohesive particles. With a size of 20 to 30 µm, these particles fluidize under very difficult to achieve conditions, and may require the application of an external force, such as mechanical agitation.

Group D The particles in this region are above 600 µm and typically have high particle densities. Fluidization of this group requires very high fluid energies and is typically associated with high levels of abrasion. Drying grains and peas, roasting coffee beans, gasifying coals, and some roasting metal ores are such solids, and they are usually processed in shallow beds or in the spouting mode.

Typically, pressurized gas or liquid enters the fluidized bed vessel through numerous holes via a plate known as a distributor plate, located at the bottom of the fluidized bed. The fluid flows upward through the bed, causing the solid particles to be suspended. If the inlet fluid is disabled the bed may settle or pack onto the plate.

Figure 1.1: Oldest power station utilizing circular fluidized bed technology, in Lünen, Germany country.

1.2 PROBLEM STATEMENT

This study is about the design and simulation of the perforated plate which work like annular distributor for fluidized bed. The annular plate is design to produce swirling air flow. The factors that need to count is parameter of the plate such as thickness, diameter, number of hole and distance of each hole that are need to consider in producing of swirling motion of air flow.

1.3 OBJECTIVE

To accomplish this project, an objective was determined:

i. To design perforated plates that produced swirling air pattern.
ii. To study the characteristic of distributor plates that have contribute to swirling of air with low pressure drop.

1.4 SCOPE OF STUDY

The details about the project is,

i. Design the perforated plates (distributor)
ii. Characteristic of the plate need be considered such as thickness, diameter, number and distance of each hole.
iii. CFD Simulation of SFB by using designed perforated plates.

CHAPTER 2

LITERATURE REVIEW

2.1 GELDART CLASSIFICATION OF PARTICLES

Not every particle can be fluidized. The behavior of solid particles in fluidized bed depends mostly on their size and density. A careful observation by Geldart (1973) is shown in figure 1. There are four different types of materials categorized.

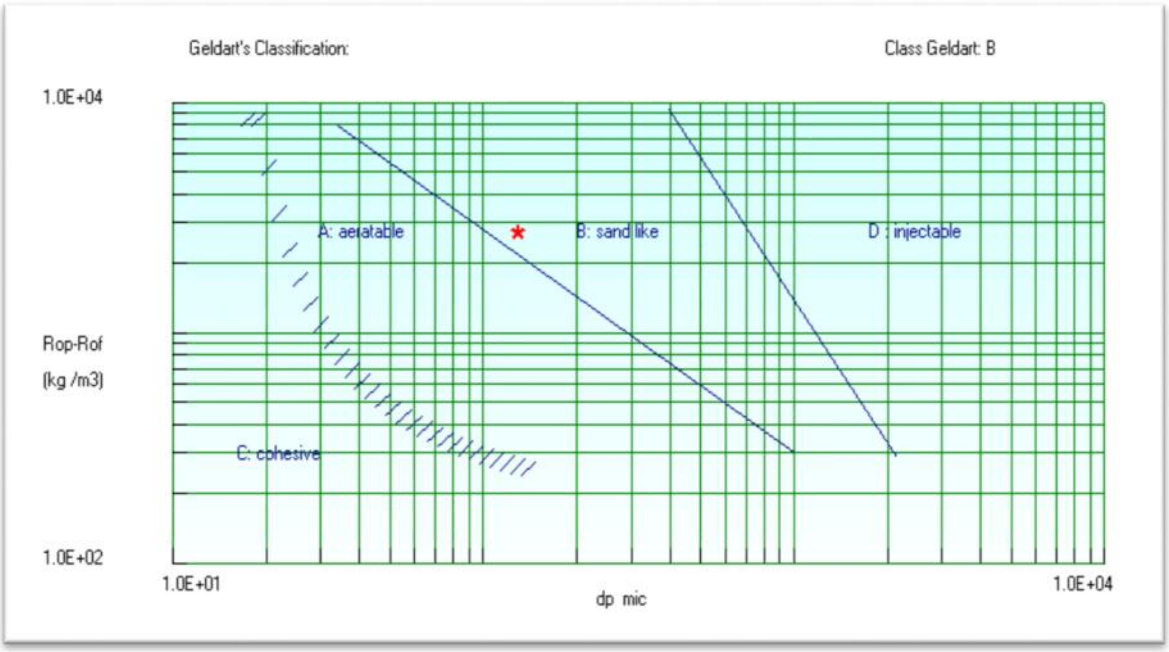

Figure 2.1: Geldart classification of particles (Geldart-1973).

Geldart type-D particles are typically large (mean size larger than 0.6 mm) and denser than other categories. They require higher velocities to fluidize the bed than other categories, resulting in the gas flow through the particle voids becoming transitional or turbulent. The bubbles which cause mixing of particles in the bed, now coalesce easily to form larger but fewer bubbles. Hence the Deldart type-D particles are difficult to fluidized, especially for deep beds and do not mix well [8][9] through spoutable. Despite their use in a large number of applications, especially in food and biomass processing, this type of particle, and its hydrodynamics in particular, have received rather less attention in publication. Cranfield and Geldart [10] studied the fluidization characteristic as of large particle (1-2mm) and discussed advantages of using fluidized beds of large particles for certain application. Rhodes [11] reviewed a number of research works on coarse particles in discussing his findings on turbulent fluidization. The mechanisms of gas flow and bubble characteristics of fluidized beds of coarse particles were investigated by Glickman. [12].

The present study explores the capability of a relatively new technique in fluidization; the swirling fluidization technique in fluidizing the Geldart type-D particles. The swirling fluidized bed (SFB) which is annular in shape with inclined injection of fluidizing gas is used with spherical PVC particles with diameters ranging from 3.85mm to 9.84mm and densities ranging from 840 kg/m^3 to 1200 kg/m^3. The bed was investigated for flow regimes, bed pressure drop ΔPb, minimum fluidization velocity, Umf and minimum swirling velocity U_{ms} experimentally. Various bed configurations were studied-different canter bodies (cone and cylinder) and bed weight from 0.5 kg to 2 kg for superficial velocities, Vs up to 6 m/s.

Another bed that operates using swirling fluidization technique is the swirling fluidized bed (SFD). The bed is annular type, featuring angular injection of gas and swirling motion of bed material in a circular path. The principle of operation is based on the simple fact that a horizontal motion of the bed particles. A jet of gas enters the bed at an angle Θ_b to the horizontal. Due to angular injection, the gas velocity has two components. The vertical component U_v=U sin Θ_b, causes ing of the particles. It is this lifting force that is responsible for fluidization. The horizontal component U_h=U cos Θ_b creates a swirling motion of the particles [13][14][15]. The bed particles are also likely to undergo a secondary motion in a toroid-like path and be well mixed in the radial plane.

This variant of fluidized bed provides an efficient means of contacting between gas and particles. Elutriation of particles which has been a major limiting factor in the operation of the conventional fluidized bed is reduced significantly, since the vertical component of velocity is now only a small fraction of the net gas velocity. The cyclone-like features resulting from the swirling motion of bed particle also contribute to this low elutriation. Hence it is capable in fluidizing a wide variety of shape of particles including the large ones.

2.2 THE PHENOMENON OF FLUIDIZATION

When we pass a fluid upward through a bed of fine particle at a low flow rate, fluid merely percolates through the void spaces between stationary particles. This is fixed bed. With an increase in flow rate, particles move apart and a few are seen vibrate and move about in restricted regions. This is the expended bed. At a still higher velocity, a point is reached when the particles are all just suspended in the upward flowing gas a liquid. At this point the fractional force between a particle and fluid counter balances the weight of the particles, the vertical component of the compressive force between adjacent particles disappears, and the pressure drop through any section of the bed about equals the weight of fluid and particles in that section. The bed is considered to be just fluidized and is referred to as an incipiently fluidized bed or a bed at minimum fluidization. In liquid solid systems and increase in flow rate above minimum fluidization usually result in a smooth, progressive expansion of the bed. Gross flow instabilities are damped and remain small, and large scale bubbling or heterogeneity in not observed under normal conditions. A bed such as this is called a particularly fluidized bed, a homogeneously fluidized bed, a smoothly fluidized bed, or simply a liquid fluidized bed.

Gas-solid systems generally behave in quite a different manner. With an increase in flow rate beyond minimum fluidization, large instabilities with bubbling and channeling of gas are observed. At higher flow rates agitation becomes more violent and the movement of solids becomes more vigorous. In addition, the bed does not expand much beyond its volume at minimum fluidization. Such a bed is called an aggregative fluidized bed, a heterogeneously fluidized bed, a bubbling fluidized bed, or simply a gas fluidized bed. In a few rare cases liquid-solid systems will not fluidized smoothly and gas solid systems will not bubble. At present such beds are not laboratory curiosities of theoretical interest.

Both gas and liquid fluidized beds are considered to be dense phase fluidized beds as long as there is a fairly clearly defined upper limit or surface to the end. However, at a sufficiently high fluid flow rate the terminal velocity of the solids is exceeded, the upper surface of the bed disappears, entrainment becomes appreciable and solids are carried out of the bed with the fluid stream. In this state we have a disperse, dilute, or lean-phase fluidized bed with pneumatic transport of solids.

2.3 BED BEHAVIORS

A detailed qualitative description of the bed behavior can be found in [16]. As the flow rate is increased, we come across the following regimes:

i. Bubbling
ii. Wave motion with dune formation
iii. Two – layer fluidizations
iv. Stable swirling

2.4 PRESSURE DROP CRITERIA FOR UNIFORM FLUIDIZATION

The pressure drop across a distributor is conventionally expressed as its ratio to the bed pressure drop, $\Delta Pd/\Delta Pb$. As a general rule of thumb, this ratio has been chosen [17] at 0.1 for deep beds. This distributor drop ΔPd is also suggested to be 10-12in. water column in a shallow bed [18] or generally 100 times the free expansion value [18] for uniform fluidization. The $\Delta Pd/\Delta Pb$ ratio is said [19][20] to fall in range 0.1-0.4 for uniform operation. The key problem is to select the aspect ratio corresponding to this pressure drop ratio. In a deep fluidized bed pressure drop is high and gas bypass as large bubbles or slugs which affect in turn heat and mass transfer rates. Shallow fluidized beds have low bed pressure drop. They have low transport disengaging height and high solid expansion ratio. There is insufficient time for the bubbles to grow and form slugs. High rate of heat and mass transfer takes place near the distributor. Shallow beds are used in industries for drying, cooling, waste heat recovery, peroxidation and cooling of iron and combustion of powdered coal. Hence Kwauk [21] stressed a need for intensifying research on shallow beds.

In order to ensure stable operation it is apparent that the pressure drop through the distributor should be sufficiently large so that the flow rate through it is relatively undisturbed by the bed pressure fluctuations above it.

Treated as a combination of a sudden contraction followed by a sudden enlargement, a simple drilled orifice in a distribution plate would be expected to have an overall pressure drop given by

$$\Delta H_d = 0.5\left(\frac{u^2}{2g}\right) + (\frac{u_0^2}{2g})$$

In consistent units, or

$$\frac{2g\Delta H_d}{u_0^2} = 1.5\ velocity\ heads$$

However, unless the plate is very thick compare with the orifice diameter ($\text{i.e.} \frac{d}{t} \ll 1$), the expansion loss will be influenced by flow patterns resulting from the sudden contraction of the flow on entry to the orifice.

$$\frac{2g\Delta H_d}{u_0^2} = 1/C_d^2$$

C_d is coefficient of discharge.

C_d is a weak function of the distributor free area () and d/t.taking a rough correlation as

$$C_d=0.82(d/t)^{-0.13}$$

Substitution in the above equation yields

$$\frac{2g\Delta H_d}{u_0^2} = 1.49(\frac{d}{t})^{0.26}$$

(A.E.QURESHI & D.E.CREASY 1978)

Figure presents the results obtained for pressure drop across the bed as the superficial gas velocity was increased. At relatively low superficial gas velocity, the pressure drop across the bed was approximately proportional to the superficial gas velocity. However, the pressure drop values were constant at above the minimum fluidization velocity, Umf. The consistency in pressure drop showed that the fluidizing gas stream had fully supported the weight of the whole bed in the dense phase. Thus Umf reached when the drag force of the up-wards fluidizing air equals to the bed weight. In this case, Umf was determined as 1.35 ms^{-1}. (S.M. Tasirin, S.K. Kamarudin* and A.M.A. Hweage 2008)

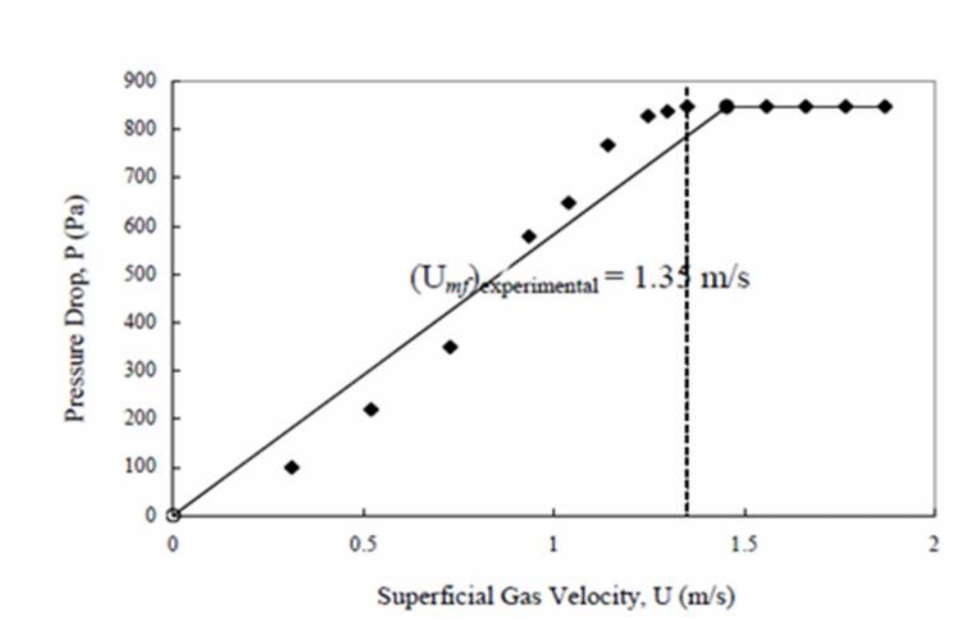

Figure 2.2: Pressure drop versus superficial gas velocity (at increasing gas flow rate) for initially mixed/segregated mixtures.

2.5 CRITICAL VELOCITY FOR UNIFORM FLUIDIZATION

Mori and Moriyama [21] attempted to relate the distributor to bed pressure drop ratio with the uniformity of fluidization and hence they linked it to the condition of no drift fluidization corresponding to last nozzle operation in a distributor. They assumed that the cross-sectional area of the fluidized bed section at the condition of no drift in fluidization is same as the total cross-sectional area of the bed and the flow through the stationary beds tends to be the same as minimum fluidization velocity. In other word a no uniformly fluidized bed is viewed to have two parts namely a fixed bed or stationary section and a fluidized bed section.

CHAPTER 3

METHODOLOGY

3.1 INTRODUCTION

In order to describe the methodology involved in this study, this chapter will be devoted to discuss the software process model which including the planning, analysis and design. The hardware and software specification that required for this project also will be discussed in this chapter. The flow chat has been plotted according to the research objectives. The first step involved is sketch out perforated plate followed by geometry simulation in Solid Work.

The hardware and software will influence the simulations. So, in this project it must run the software and hardware properly that can make a good output result of the simulations. For calculation ergun62 software is chosen as a medium of calculation parameters in testing the designing plate are working or not. After the calculation in matching a good value of parameter, next step is draw the perforated plates using the solid work and furthermore make a simulation on it to look the result. These tests were conducted in order to get the results and achieve the objectives.